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Canada Centre for Mineral and Energy Technology | Centre canadien de la technologie des minéraux et de l'énergie

ECONOMICS OF CHEMICAL METHODS FOR COPPER EXTRACTION & REFINING

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Presented at the 69th
Canadian Chemical Conference
Kingston, Ontario
June 3-5, 1985

JUL 03 1985

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MINERALS RESEARCH PROGRAM
MINERAL SCIENCES LABORATORIES
DIVISION REPORT MRP/MSL 85-85 (TR)

30 pp

MICROMEDIA

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ABSTRACT

A chloride based hydrometallurgical process for the recovery of copper and silver from copper sulfide concentrates is briefly described. This process, called the Great Central Mines (GCM) process, has been developed in Canada on a laboratory scale, and its commercial feasibility is presently being investigated. The economics of the process are compared to conventional pyrometallurgical processes for copper, and also to a sulfate based hydrometallurgical alternative. Comparative energy requirements and capital and operating costs are documented, and the effect of copper purity, and potential sulfur and iron by-product credits on overall economic viability is discussed.

RÉSUMÉ

Ce rapport contient une brève description d'un procédé hydrométallurgique à base de chlorure en vue de la récupération du cuivre et de l'argent des concentrés de sulfure cuivrique. Ce procédé, appelé procédé de la Great Central Mines (GCM), a été mis au point en laboratoire au Canada et on effectue présentement une étude de faisabilité commerciale du procédé. Le côté économique du procédé est comparé au coût des procédés pyrométallurgiques classiques pour le cuivre et au coût d'un procédé hydrométallurgique à base de sulfate envisagé comme solution de remplacement. De plus, une étude documentaire est présentée sur les besoins relatifs d'énergie et sur les frais de premier établissement et d'exploitation; l'effet de la pureté du cuivre et les avantages possibles de sous-produits du soufre et du fer sur la viabilité économique en général sont aussi discutés.

1. BACKGROUND

The Great Central Mines (GCM) hydrometallurgical copper process is a chloride based process capable of recovering high purity copper, and silver and sulfur by-products, from chalcopyrite concentrates. The recovery of a metallic iron by-product also appears feasible if economics so warrant.

Process development was initiated in 1976 when Bacon, Donaldson & Associates of Vancouver, B.C. was requested by Great Central Mines to investigate a novel copper electrowinning concept. This concept was later modified considerably by Bacon Donaldson and a complete processing system was developed and tested in 1977. A patent on the process was granted in 1979[1], and larger scale laboratory testwork was initiated. Results of this testwork were reported by Bacon Donaldson in April 1980 [2]. At the request of Great Central Mines an evaluation of the process technology and economics was carried out by CANMET in June 1980 [3]. Based on the conclusions and recommendations tabled in the CANMET report, further development work on the process was authorized by GCM and carried out at Bacon Donaldson laboratories. Results of this work were reported in November 1981 [4]. Preliminary review of these results by CANMET resulted in some further work being carried out by Bacon Donaldson and reported in April and June 1982 [5,6].

At this point, CANMET was again requested by Great Central Mines to evaluate the improved version of the GCM process and to make recommendations for commercial exploitation. A report on the CANMET evaluation was issued in November 1982 [7]. Subsequently further bench scale testwork was carried out by Bacon Donaldson to confirm the feasibility of recovering silver and producing a high purity copper product. In May 1984, these improvements in the GCM process were incorporated into new patent applications in

Canada, the United States and in several other countries. All rights to the process technology are owned by Great Central Mines.

To date nearly $1 million has been spent on the process development, including preliminary design and cost estimates for a 6 tpd pilot plant.

2. PROCESS DESCRIPTION AND CONCEPTUAL FLOWSHEET

The GCM process flowsheet is shown in Figure 1. In the process reground concentrate is leached at atmospheric pressure with acidic ferric/cupric chloride to extract at least 99% of the copper and 90% of the silver, and to convert reactive sulfides to elemental sulfur. A two-stage leach with inter-stage thickening is used to ensure that pregnant solution going to the electrowinning circuit is in the cuprous/ferrous state. After silver removal, the metal chloride solution is electrolyzed in diaphragm cells to produce a granular copper product and regenerate a portion (approximately 33%) of the ferric/cupric chloride lixiviant in the anolyte compartment. Regeneration of the remainder of the lixiviant is carried out with air in a turboaerator, prior to recycling to the oxidation stage of leaching.

The leach residue from the oxidation leach consists of gangue, pyrite and up to 65% sulfur depending on the grade of concentrate. The sulfur can be recovered by simple flotation, melting and filtering techniques to produce a saleable product.

Any gold originally present in the concentrate will remain in the leach residue.

Two stages of electrowinning are used at high current efficiency to recover about 87% of the copper entering the circuit. About 9% of the copper is recycled to leaching and the remaining 4% is recovered in a stripper cell operating at low current density to minimize hydrogen generation. The stripped solution, which is essentially copper free, becomes the bleed stream to prevent iron and impurity build-up in the process. The size of this bleed stream is dependent upon the leach conditions used, and on the concentration of iron in solution. Normal bleed stream requirements are in the range of 20 to 25% of the electrolyte flow.

A number of options can be used to remove iron and recover HCl from the bleed stream. For the GCM process the preferred option is to oxy-hydrolyze the stripped solution with air at low to medium temperatures to produce a readily disposable iron oxide product. The Lurgi [8], Stelco [9] and Pori [10] processes have all been demonstrated commercially in the steel industry for treating hydrochloric acid pickle liquors, which are very similar to the GCM strip solution. Both the USBM [11] and Bacon Donaldson have shown that a turboaerator can also be used at relatively low temperatures ($< 100°C$) to produce a discard iron oxide product and HCl for recycle to the process. At the present stage of development it is not known which of these processes will be the most cost effective for treating the GCM process ferrous chloride bleed stream. Studies to date have indicated that the lower temperature processes would be preferable if based solely on energy considerations [12]. Regardless of the option eventually chosen, water balance considerations indicate that a fractionation column for HCl upgrading will be required.

Another alternative for treating the bleed stream was investigated by Bacon Donaldson. This alternative utilizes an evaporator/crystalizer to produce hydrated ferrous chloride crystals which are then dried, pelletized and reduced with hydrogen to produce a high purity sponge iron by-product and high strength HCl for recycle to the process. This option eliminates water balance problems in the process without the need for a fractionation column. However it is a much more expensive option and is very dependent upon obtaining adequate markets and prices for the sponge iron by-product.

3. LABORATORY INVESTIGATIONS

All of the steps in the leaching, electrowinning and regeneration circuits, and two of the iron removal/recovery options have been tested at Bacon Donaldson on a laboratory scale. Processing conditions, materials of construction and material and energy balances have been reasonably well documented, and capital and operating costs have also been estimated. The main chemical reactions taking place in the GCM process are shown in Figure 2.

To date, development work on the process has been carried out on copper concentrate which is typical of those produced from British Columbia porphyry copper deposits. The composition of this concentrate is shown in Figure 3. This is a relatively high grade concentrate, with low concentrations of impurities.

Optimum leaching conditions, based on the laboratory tests, are shown in Figure 4.

Optimum electrowinning conditions and composition of electrolytes are shown in Figures 5 & 6 respectively.

Limited analyses of the copper produced under "good" and "poor" operating conditions are shown in Figure 7. To date insufficient analyses have been obtained on the copper product to define optimum conditions for producing high purity copper.

A more detailed description of the GCM hydrometallurgical copper process can be found in a recent publication (13)

4. ECONOMIC EVALUATION

Based on the laboratory test data, material and energy balances and capital and operating costs were estimated for a 70,000 stpy (63,640 tonnes/yr) plant using the GCM process.

To determine the possible effects of the various alternatives for iron removal, two versions of the GCM process have been costed, one with by-product iron recovery and one without. The costs of the latter are based primarily on USBM cost estimates for a turboaerator circuit (14). The costs of the iron recovery option are based on "Peace River Mining & Smelting" plant costs (15) and on the Bacon Donaldson testwork.

4.1 Capital Costs

For the leaching, electrowinning and regeneration circuits, equipment sizes and costs are based on the Bacon Donaldson test data using appropriate metallic and non-metallic materials of construction to meet the corrosive conditions. All leach tanks, thickeners, filters and electrowinning cells are insulated and either completely enclosed or appropriately vented to ensure containment of acid vapours. Appropriate safety factors are used in the leaching and regeneration circuits to ensure adequate retention times for continuous operation. Thickeners and filters are designed from laboratory scale settling and filtering rates, in which an anionic flocculant was added.

For iron removal using a turboaerator, equipment costs are based on the USBM cost estimates (14) but include a fractionation column for HCl upgrading. For the iron recovery option equipment includes an evaporator/crystallizer, filter, flash drier, reduction reactor, HCl absorber and heat exchangers to minimize energy requirements.

Capital cost estimates for the GCM process, with and without iron recovery are shown in Figure 8. The figures indicate that the capital costs of the iron recovery option will be about $41 million higher than the costs without iron recovery ($114 M vs $155 M).

4.2 Operating Costs

Operating cost estimates for the GCM process, with and without iron recovery are shown in Figure 9. Utility, labour and reagent costs are based on a

British Columbia location. The following unit costs were used to generate the cost estimates:

Power:	2¢/Kwh
Natural Gas:	\$3/1000 ft^3
Direct Labour:	\$15/hr (without overhead)
HCl:	\$124/t (20° Baume)
Hydrogen:	\$2.60/1000 ft^3 (16)

Make-up HCl requirements for the process are based on a 2% loss for the iron recovery option and a 4% loss without iron recovery. Natural gas requirements are based on the energy requirements shown in Figure 10 and an assumed 80% boiler efficiency for steam generation.

The operating costs in Figure 9 indicate that the iron recovery option will increase costs by about 1.5¢/lb of copper. However if the iron by-product can be sold for \$100/ton the net result will be an operating cost savings of about 2.8¢/lb or about \$3.9 million/yr. This would not appear to be adequate to provide an acceptable return on investment on the additional \$41 million required for the iron recovery plant.

4.3 Energy Requirements

A breakdown of the energy requirements for the GCM process with and without iron recovery is shown in Figure 10.

Although the thermal energy requirements for the iron recovery option are high, they are still estimated to be lower than those for iron disposal because of the

recuperation to minimize total energy requirements is also difficult, at least for this low temperature iron disposal option. It should be possible however to utilize the heat generated in the regeneration circuit to reheat the leach liquor and reduce the energy requirements in the leach circuit to about 100 Btu/lb Cu. Total thermal energy requirements therefore could be as low as 7880 Btu/lb Cu with iron recovery, and 8900 Btu/lb Cu without iron recovery. With a very efficient system it may be possible to reduce energy requirements in the latter to about the same as those for iron recovery.

4.4 By-Product Sulfur Recovery

Although the feasibility and cost of recovering sulfur from the GCM process leach residues has not yet been demonstrated, there is little doubt that a saleable sulfur by-product can be produced by relatively simple flotation and hot filtration techniques, as Cominco has already demonstrated in its zinc pressure leach plant at Trail, B.C. This sulfur by-product is at least double the value of sulfuric acid and can be shipped at least three times as far for the same cost in terms of equivalent sulfur. For the GCM process, capital and operating costs for sulfur production should be considerably lower than costs for sulfuric acid production from copper smelters, since all of the sulfur is produced directly in the leach process and can be recovered by relatively inexpensive techniques. Costs will also be less than in the Sherritt Cominco (S-C) copper process [17], where only about half the sulfur is produced directly in the leach, with the remainder produced in a Claus plant. Estimated sulfur recovery in the GCM process is at least 80% versus

about 60% in the S.C. process. Based on $100/tonne selling price, netbacks from sulfur sales, even from inland locations in Canada, should provide additional revenues to the GCM process of 1.5 to 3.5¢/lb of copper (depending on distance to market), after all production costs have been paid for.

5. COMPARISON WITH ALTERNATE COPPER PROCESSES

5.1 Energy Considerations

A comparison of the GCM process energy requirements with modern pyrometallurgical copper smelting processes [18] and the S-C process [17,19] is shown in Figure 11. The figures indicate that the electrical energy requirements for the GCM process are in the same range as the modern smelting processes using oxygen, and are less than 1/3 of the S-C process requirements. Although the thermal energy requirements for both the GCM and S-C processes are much higher than the modern pyro processes, the requirements are primarily in the form of low grade steam, which can be obtained from low cost coal, whereas the smelters for the most part must use high cost fuels. For the hydro processes therefore thermal energy costs may be as little as 1.3¢/lb copper higher than the modern pyro processes costs, if the cheapest fuel is used in both cases, (see Figure 12). For the GCM process total average energy costs should be less than 1¢/lb of copper higher than the pyro process costs and at least 2¢/lb of copper less than the S-C process costs.

5.2 Capital & Operating Costs

Capital and operating costs for other copper extraction refining processes are shown in Figure 13 for comparison with the GCM process. The figures indicate that the GCM process capital costs will be considerably lower than the S-C process, and could be 50% lower than modern pyrometallurgical processes for copper. This capital cost differential will increase at lower plant capacities because hydrometallurgical processes in general do not have substantial economies of scale.

Operating costs also appear to be competitive with the other processes. The low power requirements alone will provide an advantage over the S-C process of at least 2¢/lb Cu. Thermal energy requirements however will still be much higher for the GCM process than for the new energy efficient pyrometallurgical processes. Operating costs therefore, due to energy considerations, are likely to favour the pyro processes by 1 to 3¢/lb of copper. However, as shown in Figure 13, sulfur by-product credits alone could more than off-set this operating cost differential.

Total costs, when capital charges amortized at 10% over 20 years are included, indicate an advantage to the GCM process of 6 to 15¢/lb of copper. This cost advantage however is very dependent on producing a high purity copper product, equivalent to electrorefined copper.

6. CONCLUDING REMARKS

The GCM hydrometallurgical copper process is a potentially attractive alternative for copper production from copper concentrates, particularly in isolated inland locations where sulfuric acid cannot be readily marketed, and SO_2 emissions cannot be tolerated. The process has lower capital costs and competitive operating costs when compared to alternate hydrometallurgical or pyrometallurgical technologies. It operates at atmospheric pressure and at reasonable temperatures.

As with most hydrometallurgical technologies the process can be economically viable at relatively low throughputs and has the flexibility to maintain high copper and silver recoveries from a variety of low or high grade copper concentrates. Since the process converts sulfides to elemental sulfur and does not produce SO_2, there are no production constraints related to sulfuric acid markets or SO_2 emissions. The elemental sulfur produced can either be recovered and sold if markets exist, or stockpiled or disposed in tailings with no detrimental effect on the environment.

The process has the potential to produce high purity copper directly, and also recover a high grade silver by-product. A saleable (copper free) metallic iron by-product can also be produced if economics so warrant.

Presently, Hudson Bay Mining & Smelting (HBMS), in co-operation with Great Central Mines, Bacon Donaldson & Canmet, is carrying out further laboratory scale development work on the GCM process to determine the technical and economic viability of the process for application to HBMS copper concentrates. This work is being carried out under the auspices of the Canada/Manitoba Mineral Development Agreement, and includes investigations on iron removal and disposal options, and on the recovery of sulfur and gold from process leach residues.

If the results of this testwork are successful, and particularly if a high purity copper product can be produced, there is little doubt that this chemical extraction process will be a very attractive option for the future production of copper in Canada and elsewhere in the world.

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FIGURE 1

GREAT CENTRAL MINES

HYDROMETALLURGICAL COPPER PROCESS

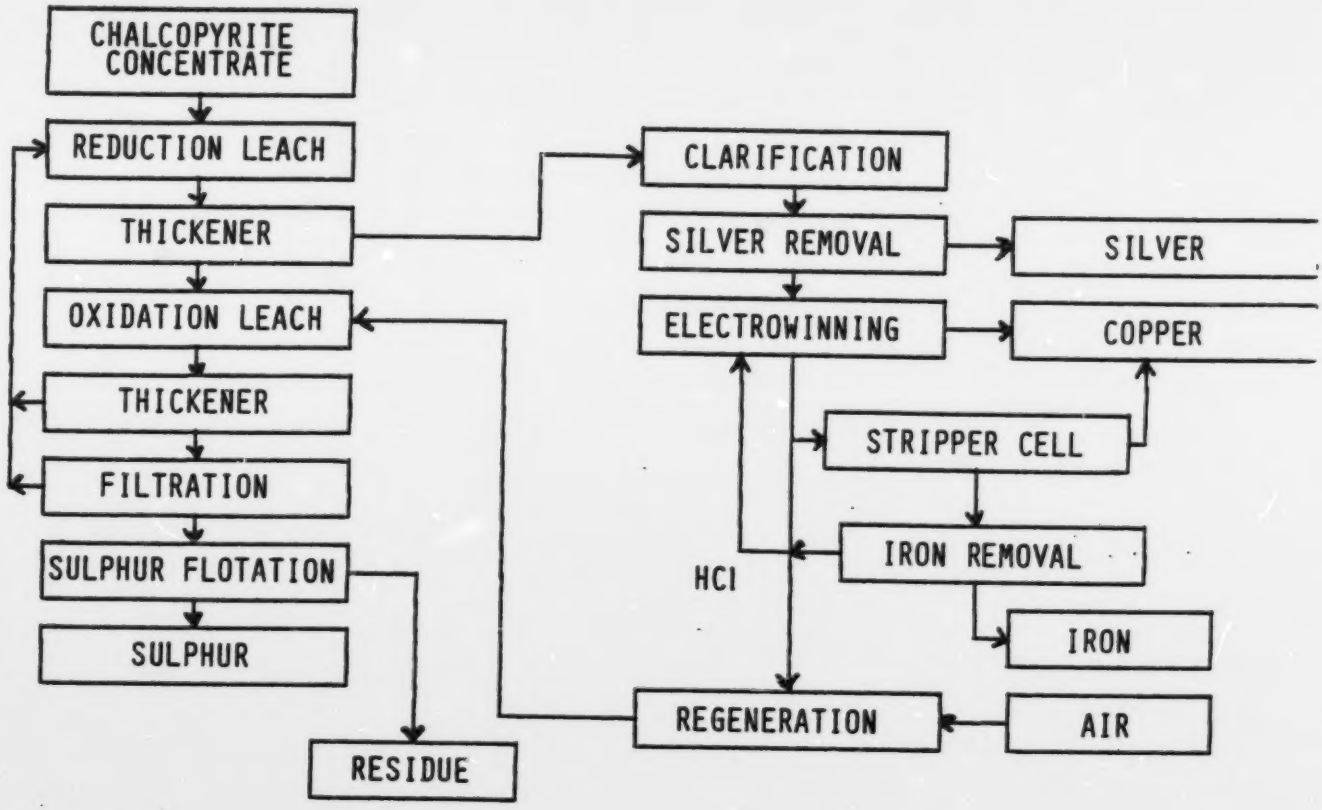


FIGURE 2

CHEMICAL REACTIONS

LEACHING

$3\ FECL_3 + CUFES_2 \rightarrow CUCL + 4FECL_2 + 2S^\circ$

ELECTROWINNING

CATHOLYTE: $CUCL \rightarrow CU^\circ + CL'$

ANOLYTE: $FECL_2 + HCL \rightarrow FECL_3 + H^+$
$CUCL + HCL \rightarrow CUCL_2 + H^+$

REGENERATION

$2FECL_2 + 0.5O_2 + 2HCL \rightarrow 2FECL_3 + H_2O$

OXYHYDROLYSIS

$4FECL_2 + O_2 + 4H_2O \rightarrow 2FE_2O_3 + 8HCL$

FIGURE 3

CONCENTRATE ANALYSIS

28.1%	CU
28.22%	FE
30.91%	S
0.03%	AS
0.013%	SB
0.011%	ZN
0.010 OZ/TON	AU
0.68 OZ/TON	AG

FIGURE 4

LEACHING CONDITIONS

CONC. GRIND: 100% MINUS 325 MESH

TEMP. : 100° - 105°C

% SOLIDS : 10% (1ST STAGE) 5% (2ND STAGE)

RETENTION TIME (BATCH): 0.5 HOURS (1ST STAGE)
3.0 HOURS (2ND STAGE)

COPPER EXTRACTION: 99% MINIMUM

SILVER EXTRACTION: 92% MINIMUM

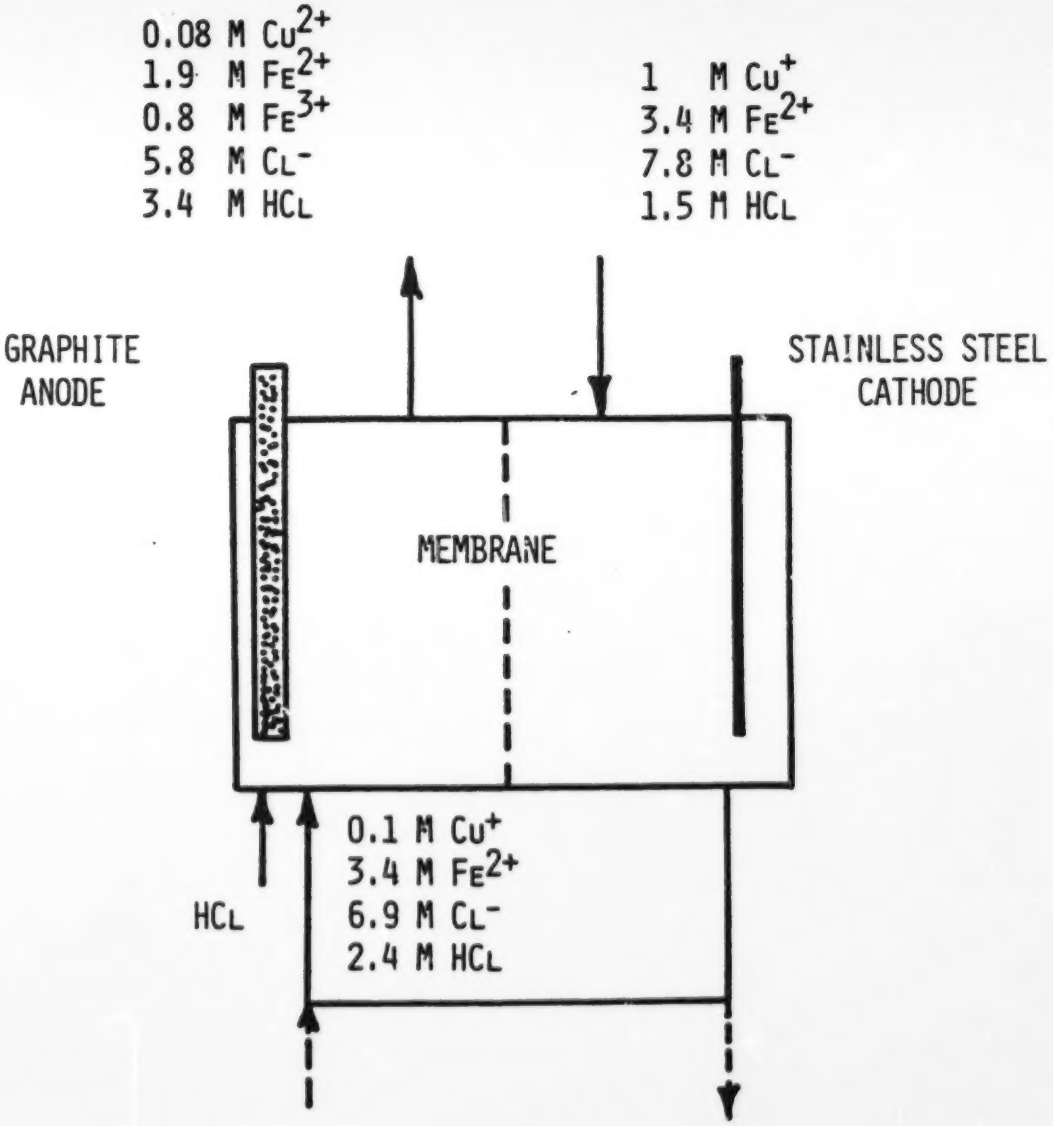
SOLUTION CONCENTRATIONS:	INITIAL	FINAL
COPPER (GPL)	5(CU+2)	60 (CU+)
IRON (GPL)	150(FE+3)	190(FE+2)

FIGURE 5

ELECTROWINNING CONDITIONS

TEMPERATURE (°C)	80
COPPER DEPOSITION RATE ($G/FT^2/MIN$) (AV)	0.93
POWER REQ'D (KWH/LB CU) (AV)	0.30
VOLTAGE (AV)	1.48
CURRENT DENSITY (A/FT^2) (AV)	24
CURRENT EFFICIENCY (%) (AV)	96
% COPPER RECOVERED PER PASS	91
% COPPER RECOVERED AT HIGH C.E.	87
% COPPER RECOVERED IN AT LOW C.E. (STRIPPER CELL)	4
C.E. IN STRIPPER CELL (%)	63

FIGURE 6



COMPOSITION OF ELECTROWINNING SOLUTIONS.

FIGURE 7

PRODUCT ANALYSIS (PPM)

OPERATING CONDITIONS	AS	BI	SB	SE	AG
GOOD	12	0.8	0.8	0.8	3
POOR	33-84	5-16	5-8	1	11

FIGURE 8

GCM PROCESS
CAPITAL COST ESTIMATES

(BASED ON 70,000 STPY CU)

	MID 1984 $ 1000
FEED PREPARATION:	2,600
LEACH CIRCUIT:	7,100
ELECTROWINNING CIRCUIT:	19,400
REGENERATION CIRCUIT:	2,900
UTILITIES & INFRASTRUCTURE	21,300
DIRECT PLANT COSTS	53,300
ENGINEERING AND FEES	18,600
CONTINGENCY	16,000
TOTAL SECTION COSTS	87,900
STEAM PLANT (TURNKEY)	3,400
IRON RECOVERY PLANT (TURNKEY)	64,200
TOTAL GCM PROCESS PLANT COSTS (WITH IRON RECOVERY)	155,500
GCM PROCESS COSTS (WITHOUT IRON RECOVERY)	
TOTAL SECTION COSTS	87,900
STEAM PLANT (TURNKEY)	3,800
IRON RESIDUE DISPOSAL PLANT (TURNKEY)	22,500
TOTAL GCM PROCESS COSTS (WITHOUT IRON RECOVERY)	114,200

FIGURE 9

GCM PROCESS OPERATING COST ESTIMATES

	C/LB CU	
	WITH IRON RECOVERY	WITHOUT IRON RECOVERY
DIRECT COSTS		
REAGENTS:		
GRINDING MEDIA	0.2	0.2
HCl	0.41	0.82
H_2 (0.032 LB/LB)	1.5	-
	2.11	1.02
UTILITIES:		
ELECTROWINNING POWER (0.33KWH/LB)	0.66	0.66
OTHER POWER (0.25 KWH/LB)	0.50	0.50
NATURAL GAS	3.27	3.84
WATER (PROCESS AND RAW)	0.10	0.10
	4.53	5.10
DIRECT LABOUR:		
LABOUR	3.25	3.25
SUPERVISION (20% OF LABOUR)	0.65	0.65
	3.90	3.90
MAINTENANCE:		
LABOUR	1.20	1.20
SUPERVISION (25% OF LABOUR)	0.30	0.30
MATERIALS (1% OF CAPITAL)	1.11	0.81
	2.61	2.31
PAYROLL OVERHEAD:		
35% OF LABOUR	1.90	1.90
TOTAL DIRECT COSTS:	15.05	14.23
INDIRECT COSTS:		
35% OF DIRECT LABOUR AND MAINTENANCE	2.28	2.17
FIXED COSTS:		
TAXES (1% OF PLANT COST)	1.11	0.81
INSURANCE (1% OF PLANT COST)	1.11	0.81
	2.22	1.62
TOTAL OPERATING COSTS (EXCLUDING AMORTIZATION)	19.55	18.02
CREDIT FOR IRON ($100/TON)	(4.3)	-
NET OPERATING COSTS	15.25	18.02

FIGURE 10

GCM PROCESS ENERGY REQUIREMENTS

WITH IRON RECOVERY BTU/LB CU		PROCESS STAGE	WITHOUT IRON RECOVERY BTU/LB CU	
1,440		LEACHING	1,440	
0	(-1340)(1)	REGENERATION	0	(-1340)
-		IRON OXIDATION/DISPOSAL	8,800	
-		IRON RECOVERY	-	
3,200		CRYSTALLIZATION	-	
1,600		DRYING	-	
2,980		H_2 REDUCTION	-	
9,220	(7,880)	TOTAL THERMAL	10,240	(8,900)
KWH/LB CU			KWH/LB CU	
0.33		ELECTROWINNING (2)	0.33	
0.25		OTHER ELECTRICAL	0.25	
0.58		TOTAL ELECTRICAL	0.58	

(1) REGENERATION IS EXOTHERMIC AND RESULTS IN THE EQUIVALENT OF 1,340 BTU/LB OF COPPER BEING RELEASES.

(2) INCLUDES A 10% SAFETY FACTOR FOR COMMERCIAL SCALE OPERATION.

FIGURE 11

ENERGY CONSUMPTION IN
COPPER EXTRACTION/REFINING PROCESSES

PROCESS	ELECTRICAL KWH/TONNE CU	THERMAL 10^6 BTU/TONNE CU	
		WITHOUT STEAM CREDITS	WITH STEAM CREDITS
MODERN PYRO PROCESSES USING OXYGEN (1, 2)	1,100-1,280	3.0-8.5	2.0-5.5
S-C PROCESS	3,840-4,200	16.5-25.0	
GCM PROCESS	1,200-1,300	17-22.5	

(1) INCLUDES INCO, NORANDA, MITSUBISHI AND OUTOKUMPU PROCESSES.

(2) INCLUDES DRYING, SMELTING, CONVERTING, ANODE FURNACE REFINING, CASTING, ELECTROREFINING, OXYGEN AND ACID PLANTS.

FIGURE 12

COMPARATIVE ENERGY COSTS
OF COPPER EXTRACTION/REFINING PROCESSES

C/LB CU

	ELECTRICAL (1)	THERMAL (2)	TOTALS RANGE	AV
MODERN PYRO PROCESSES	1.0-1.2	0.3-2.7	1.3-3.9	2.6
S-C PROCESS	3.5-3.8	1.5-3.4	5.0-7.2	6.1
GCM PROCESS	1.1-1.2	1.6-3.1	2.7-4.3	3.5

(1) BASED ON 2C/KWH.

(2) BASED ON AVERAGE COAL, NATURAL GAS AND HEAVY FUEL OIL PRICES IN B.C. ASSUMES NATURAL GAS OR HEAVY FUEL OIL FOR SMELTERS AND NATURAL GAS OR COAL FOR THE HYDRO PROCESSES AT 100% EFFICIENCY.

FIGURE 13

CAPITAL AND OPERATING COSTS OF COPPER EXTRACTION REFINING PROCESSES

	PYRO PROCESSES		HYDRO PROCESSES		
	MODERN(1) SMELTERS	OLD SMELTERS	S-C PROCESS	GCM PROCESS (2)	(3)
PLANT CAPACITY 10^3 TONNES/YR	65-180	70-200	75	65	65
CAPITAL COST $/A TONNE CU	3,700-5,000	-	3,300	1,800	2450
OPERATING COST C/LB CU	16-20	21-29	19-21	18	19.5
BY-PRODUCT CREDIT (4) C/LB CU SULFUR	-	-	1.0-3.0	1.5-3.5	1.5-3.5
IRON				-	4.0
*CAPITAL CHARGES C/LB CU	11-15	-	10	5.5	7.5
TOTAL COSTS C/LB CU	27-35	-	26-30	20-22	19.5-21.5

* BASED ON 20 YEAR AMORTIZATION AT 10%.

(1) INCLUDES SMELTING, REFINING AND ACID PRODUCTION.
(2) WITHOUT IRON RECOVERY. (3) WITH IRON RECOVERY.
(4) ASSUMES NO CREDIT OR LOSS FOR SULFURIC ACID FROM SMELTERS.